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HYDRODYNAMICS AND MASS TRANSFER IN A
DRAFT TUBE GAS-LIQUID-SOLID SPOUTED BED

DISSERTATION

Presented in Partial Fulfillment of the Requirements for
the Degree Doctor of Philosophy in the Graduate
School of The Ohio State University

by

Shyh-Jye Hwang

*****

The Ohio State University
1985

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NOMENCLATURE

\[ A \] solid-liquid interfacial area, \( \text{cm}^2 \)

\[ A_D \] cross-sectional area of the draft tube, \( \text{cm}^2 \)

\[ A_{g i}, i=1,6 \] area for gas flow, \( \text{cm}^2 \)

\[ A_{i}, i=1,6 \] total area, \( \text{cm}^2 \)

\[ A_{li}, i=1,6 \] area for liquid flow, \( \text{cm}^2 \)

\[ A_{si}, i=1,6 \] area for solid flow, \( \text{cm}^2 \)

\[ A_T \] cross-sectional area of the column, \( \text{cm}^2 \)

\[ C \] benzoic acid concentration, g/cc

\[ C_{in} \] benzoic acid concentration at the spouted bed inlet, g/cc

\[ C_{out} \] benzoic acid concentration at the spouted bed outlet, g/cc

\[ C_{sat} \] saturated benzoic acid concentration, g/cc

\[ D \] molecular diffusivity of benzoic acid in water, \( \text{cm}^2/\text{sec} \)

\[ D_c \] column diameter, cm

\[ D_D \] draft tube diameter, cm

\[ D_e \] equivalent diameter of the annular region, cm

\[ d_b \] bubble diameter, cm

\[ d_p \] particle diameter or equivalent diameter, cm

\[ d_{b\bar{a}} \] average bubble size for the distribution based on the number frequency of bubbles, cm

\[ d_{v\bar{a}} \] average bubble size for the distribution based on the volume frequency of bubbles, cm
**F**  total friction force, dyne, or liquid flow rate, cc/sec

**F_A**  friction force in the annular region, dyne

**F_B**  friction force in the below-draft-tube region, dyne

**F_D**  friction force in the draft tube region, dyne

**F_T**  friction force in the above-draft-tube region, dyne

**f**  friction factor, defined by Eqn. (10)

**G**  dimensionless group defined in Figure 5.5 for a draft tube liquid-solid spouted bed

**G'**  dimensionless group defined in Figure 5.5 for a liquid-solid fluidized bed

**Ga**  Galileo number, defined as $d^3 \rho_f^2 g/\mu_L^2$

**g**  gravitational acceleration, cm/sec$^2$

**H**  downward distance from the top of the draft tube representing the bubble penetration depth in the annular region, cm

**K**  sudden contraction-loss coefficient

**k_2**  solid-liquid mass transfer coefficient for a draft tube liquid-solid spouted bed, cm/sec

**k_2'**  solid-liquid mass transfer coefficient for a liquid-solid fluidized bed, cm/sec

**k_3**  solid-liquid mass transfer coefficient for a draft tube gas-liquid-solid spouted bed, cm/sec

**k_3'**  solid-liquid mass transfer coefficient for a gas-liquid-solid fluidized bed, cm/sec

**L**  downward distance from the upper retaining grid, cm

**L_1**  distance from the top of the distributor to the upper retaining grid, cm

**L_B**  bottom spacing, cm

**L_D**  height of the draft tube, cm

**L_T**  top spacing, cm

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density number, defined as \((p_s - p_l)/p_l\)

the area ratio of the annular region to the mid-outer tube of the injector

\(P_i, i=1,6\) local pressure, dyne/cm²

\(\Delta P\) total pressure loss in the bed, dyne/cm²

\(\Delta P_{Ah}\) pressure drop in the annular region due to hydrostatic head, dyne/cm²

\(\Delta P_{Da}\) pressure drop in the draft tube region due to acceleration of fluid, dyne/cm²

\(\Delta P_{Df}\) pressure drop in the draft tube region due to friction, dyne/cm²

\(\Delta P_{Dh}\) pressure drop in the draft tube region does to hydrostatic head, dyne/cm²

\(\Delta P_f\) frictional pressure loss in the bed, dyne/cm²

\(Q_{\text{app}}\) apparent liquid circulation rate, l/sec

\(R(d_b)\) dimensional bubble size distribution function

\(R(\xi)\) dimensionless bubble size distribution function

\((Re)_A\) Reynolds number in the annular region, defined as \(D_eU_A\rho_l/\mu_l\)

\((Re)_c\) apparent liquid circulation Reynolds number defined as \(D_cU_c\rho_l/\mu_l\)

\(Re_g\) gas Reynolds number, defined as \(d_pU_g\rho_g/\mu_g\)

\(Re_l\) liquid Reynolds number, defined as \(d_pU_l\rho_l/\mu_l\)

\(r_h\) hydraulic radius, cm

\(r_i\) radius of the inner tube of the injector, cm

\(r_o\) radius of the outer tube of the injector, cm

\(Sc\) Schmidt number, defined as \(\mu_l/\rho_l D\)

\(Sh_{2,2}\) Sherwood number for a draft tube liquid-solid spouted bed, defined as \(k_{2,d_p}/D\)

\(Sh_{2,2}'\) Sherwood number for a liquid-solid fluidized bed, defined
$\text{Sh}_3$ Sherwood number for a draft tube gas-liquid-solid spouted bed, defined as $k_2^\prime \frac{d_p}{D}$

$\text{Sh}_3'$ Sherwood number for a gas-liquid-solid fluidized bed, defined as $k_3 \frac{d_p}{D}$

$\bar{t}$ average time for a tracer particle to make a single loop, sec

$t_A$ liquid circulation time in the annular region, sec

$t_B$ liquid circulation time in the below-draft-tube region, sec

$t_D$ liquid circulation time in the draft tube region, sec

$t_T$ liquid circulation time in the above-draft-tube region, sec

$t_{\text{Total}}$ total liquid circulation time, sec

$U_l$ liquid velocity from the injector, cm/sec

$U_A$ liquid velocity in the annular region based on the cross-sectional area of the annular region, cm/sec

$U_B$ liquid velocity in the annular region based on the cross-sectional area of mid-outer tube of the injector, cm/sec

$U_C$ apparent liquid circulation velocity, cm/sec

$U_D$ liquid velocity in the draft tube region based on the cross-sectional area of the draft tube, cm/sec

$U_g$ superficial gas velocity based on the cross-sectional area of the column, cm/sec

$U_{g_i, i=1,6}$ linear gas velocity, cm/sec

$U_L$ superficial liquid velocity based on the cross-sectional area of the column, cm/sec

$U_{g_i, i=1,6}$ linear liquid velocity, cm/sec

$U_{sl, i=1,6}$ linear solid velocity, cm/sec

$V_A$ total volume in the annular region, ml
\( V_B \) total volume in the below-draft-tube region, ml
\( V_D \) total volume in the draft tube region, ml
\( V_L \) volume of liquid in the bed, ml
\( V_S \) solids loading, ml
\( V_T \) total volume in the above-draft-tube region, ml

Greek Letters

\( \rho_g \) gas density, g/cc
\( \rho_L \) liquid density, g/cc
\( \rho_m \) mixture density, g/cc
\( \rho_s \) solid density, g/cc
\( \varepsilon_g \) gas holdup in the bed
\( \varepsilon_{gA} \) gas holdup in the annular region
\( \varepsilon_{gB} \) gas holdup in the below-draft-tube region
\( \varepsilon_{gD} \) gas holdup in the draft tube region
\( \varepsilon_{gT} \) gas holdup in the above-draft-tube region
\( \varepsilon_l \) liquid holdup in the bed
\( \varepsilon_{lA} \) liquid holdup in the annular region
\( \varepsilon_{lB} \) liquid holdup in the below-draft-tube region
\( \varepsilon_{lD} \) liquid holdup in the draft tube region
\( \varepsilon_{lT} \) liquid holdup in the above-draft-tube region
\( \varepsilon_s \) solid holdup in the bed
\( \varepsilon_{SA} \) solid holdup in the annular region
\( \varepsilon_{SB} \) solid holdup in the below-draft-tube region
\( \varepsilon_{SD} \) solid holdup in the draft tube region
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<td>$\varepsilon_{ST}$</td>
<td>solid holdup in the above-draft-tube region</td>
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<td>$\alpha$</td>
<td>angle of the conical section</td>
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<tr>
<td>$\xi$</td>
<td>dimensionless bubble diameter</td>
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<tr>
<td>$\zeta$</td>
<td>gradual contraction-loss coefficient</td>
</tr>
<tr>
<td>$\mu_g$</td>
<td>gas viscosity, g/cm/sec</td>
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<tr>
<td>$\mu_l$</td>
<td>liquid viscosity, g/cm/sec</td>
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<td>$\phi$</td>
<td>sphericity</td>
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ABSTRACT

Experiments were conducted using various types of solid particles to investigate the hydrodynamic properties of a draft tube gas-liquid-solid spouted bed. The hydrodynamic properties under study include flow modes, pressure profile and pressure drop, bubble penetration depth, overall gas holdup, apparent liquid circulation rate, and bubble size distribution. Three flow modes were classified: a packed bed mode, a fluidized bed mode, and a circulated bed mode. It was found that the friction factor accounting for the friction loss in the bed varies linearly on a logarithmic scale with the Reynolds number defined based on the apparent liquid circulation rate. The bubble penetration depth in the annular region, overall gas holdup, and apparent liquid circulation rate increase with an increase in gas or liquid velocity. At high gas flow conditions, an optimal solids loading exists which yields a maximum apparent liquid circulation rate. A model was proposed to describe the liquid circulation behavior in the draft tube gas-liquid-solid spouted bed. The average bubble size in the draft tube region is higher than that in the annular region for both the dispersed bubble regime and the coalesced bubble regime in the draft tube region.

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The optimal design of a draft tube gas-liquid-solid spouted bed was evaluated by considering four design variables including draft tube diameter, gas-liquid injector, top (above draft tube) spacing, and bottom (below draft tube) spacing. The effects of these design variables on the overall gas holdup were comprehensively examined for various gas and liquid velocities and solid particle properties. The diameter of the outer column of the draft tube gas-liquid-solid spouted bed was fixed at 15.24 cm. The overall gas holdup increases with an increase in gas or liquid velocity. It was found that among three diameters of draft tubes considered (7.62, 10.16, 12.7 cm), the 7.62 cm ID draft tube has the highest overall gas holdup. Of the two liquid velocities considered (2.28 cm/sec and 3.41 cm/sec), a higher overall gas holdup is obtained at a higher liquid velocity for each of the four bottom spacings considered (0.44, 1.27, 2.54, 5.08 cm). Among these four bottom spacings, the 1.27 cm spacing has the highest overall gas holdup. The overall gas holdup increases with decreasing top spacing. 2.5 mm nylon balls have the highest overall gas holdup, followed by 4 mm gass beads. 3 mm glass beads have the lowest overall gas holdup. The concentric tube gas-liquid injector with the smallest pore size gas distributor plate placed on the top of the outer tube gives rise to the highest overall gas holdup.

Dissolution of benzoic acid pellets into water was used to obtain the solid-liquid mass transfer coefficient in a draft tube gas-liquid-solid spouted bed. The in-bed benzoic acid concentration was examined to justify the use of the continuous flow stirred tank.
reactor (CSTR) model in the evaluation of the mass transfer coefficient. Effects of operating parameters including gas and liquid flow rates, solid particle size, and solids loading on the solid-liquid mass transfer coefficient were analyzed. An empirical correlation was proposed to correlate the solid-liquid mass transfer coefficient in the draft tube liquid-solid and gas-liquid-solid spouted beds. Comparisons of the mass transfer coefficient among the liquid-solid fluidized bed, gas-liquid-solid fluidized bed, and draft tube gas-liquid-solid spouted bed were also made.
CHAPTER 1

INTRODUCTION

The gas-liquid-solid fluidized bed has emerged in recent years as one of the most promising device for gas-liquid-solid operation. Such a device is of considerable industrial importance as evidenced by its wide use for chemical, petrochemical, and biochemical processing (e.g., Van Driesen and Stewart, 1964; Eccles and DeVax, 1981; Hirata et al., 1982). Work in the gas-liquid-solid fluidized bed where liquid is the continuous phase is conducted mainly with the cocurrent upward flow of the gas and liquid (e.g., Efremov and Vakhrushev, 1970; Epstein and Nicks, 1976; Begovich and Watson, 1978). While recent emphasis has been placed on the optimal design and operation of the gas-liquid-solid fluidized bed, it is highly desirable to examine the characteristics of different or new modes of gas-liquid-solid operation with liquid as the continuous phase. Such modes of gas-liquid-solid fluidized bed operation are exemplified by inverse gas-liquid-solid fluidization (Werner and Shugerl, 1972; Fan et al., 1982a, 1982b; Chern et al., 1983), constrained gas-liquid-solid fluidization (Chern et al., 1983), and gas-liquid-solid semifluidization (Edwards, 1981; Edwards, 1981).
Beaver et al., 1982; Chern et al., 1982). The present study addresses a new mode of gas-liquid-solid fluidization characterized by draft tube gas-liquid-solid spouted bed operation.

In the draft tube gas-liquid-solid spouted bed, a draft tube is coaxially situated inside the bed with a gas and liquid injector located under the draft tube. The draft tube serves to provide bulk circulation of gas, liquid, and solids between the draft tube and annulus, and thereby achieve intimate contact among gas, liquid, and solid phases in the bed. The operation of the draft tube gas-liquid-solid spouted bed in the absence of solid particles characterizes an internal loop airlift reactor.

The airlift reactor has been used extensively for aerobic fermentation and bio-oxidation of wastewater in wastewater treatment processes because of its high mass transfer rate between gas and liquid phases and low power consumption. The most commonly known airlift reactor consists of two concentric cylinders referred to as the outer column and the inner column (or draft tube) (e.g., Hatch, 1973; Malfait et al., 1981). Gas can be sparged into the base of either the draft tube region or the annular region. This results in a pressure difference between the draft tube region and the annular region at the top and bottom levels of the draft tube. This pressure difference constitutes a driving force for liquid circulation between the draft tube and annular regions. Gas is entrained by the circulating liquid to the annular or draft tube region. Other designs of airlift reactors can also be found in the literature. They
include: external loop airlift reactor (Weiland and Onken, 1981a, 1981b), split cylinder airlift reactor (Kubota et al., 1978; Orazem and Erickson, 1979), thin channel rectangular airlift reactor (Gasner, 1974), double draft tube airlift reactor (Margaritis and Sheppard, 1981), and multi-stage airlift reactor (Orazem and Erickson, 1979; Mao et al., 1982).

Extensive study of the airlift reactor has been made on the effects of the geometry of the reactor, gas and liquid flow rates, and power input on the mass transfer rate between the gas and liquid phases (e.g., Hatch, 1973; Malfait et al., 1981; Weiland and Onken, 1981a, 1981b), and on the fundamental hydrodynamic characteristics (e.g., Kubota et al., 1978; Keitel and Onken, 1981; Fields and Slater, 1983). However, little is known regarding the airlift reactor when operated in the presence of solid particles (Fan et al., 1984a, 1984b). In a gas-liquid-solid system with gas and liquid introduced into the draft tube through a gas-liquid tube injector, a liquid spout (jet) containing gas bubbles and solid particles forms above the injector. The size of the spout varies with the liquid velocity and the injector design. A system of this nature can best be described as a draft tube gas-liquid-solid spouted bed.

This dissertation is divided into seven chapters. Chapter Two is a literature review. Chapter Three presents the hydrodynamic properties of the draft tube gas-liquid-solid spouted bed. Specifically, the hydrodynamic properties including flow modes, pressure profile and pressure drop, bubble penetration depth in the
annular region, overall gas holdup, apparent liquid circulation rate, and bubble size distribution in the draft tube and annular regions are explored. Effects of gas and liquid velocities, solids loading, and physical properties of solid particles on these hydrodynamic properties are extensively analyzed.

Chapter Four presents some design considerations in a draft tube gas-liquid-solid spouted bed. The overall gas holdup is chosen as a convenient dependent variable on which the optimal design of a draft tube gas-liquid-solid spouted bed is based. In Chapter Five, the solid-liquid mass transfer behavior in a draft tube gas-liquid-solid spouted bed of air, dilute benzoic acid solution, and benzoic acid particles is studied. Effects of operating parameters including gas and liquid flow rates, solid particle size, and solids loading on the solid-liquid mass transfer coefficient are analyzed.

Chapter Six is the concluding remarks and Chapter Seven presents some recommendations for future study.
A. Spouted Bed

In a spouted bed, the bottom part of a cylindrical column is in a conical shape and the bed of particles is fluidized by the injection of gas and liquid upward from the lower end of the cone. The solid particles in the core area of the spouted bed are carried by the fountain and separated from the gas or liquid at the surface of the fountain. The solid particles then moved downward in the annular area. Investigations of the spouted bed have been mainly focused on the gas-solid systems (Mathur and Epstein, 1974).

Four different regimes can be observed in gas-liquid-solid spouted bed operation, namely, coalesced bubble flow regime, dispersed bubble flow regime, slug flow regime, and gas continuous flow regime. Nishikawa et al. (1976, 1977) measured the gas holdup in a cocurrent gas-liquid-solid spouted bed with the liquid as the continuous phase. Gas-liquid mass transfer in the same type of reactor was also measured by Nishikawa et al. (1977). Kono (1980) developed a cocurrent gas-liquid-solid spouted bed reactor with gas as the continuous phase.
The product of this reactor was granulated solids. A gas-liquid-solid spouted bed with the gas as the continuous phase operating under countercurrent conditions can also be utilized for gas-liquid mass transfer applications (Vukovic et al., 1974).

B. Jet Reactor

Reactors which are equipped with two phase injectors-ejectors are commonly known as jet reactors in the energy producing industry. These two phase injectors-ejectors are devices which utilize the kinetic energy of liquid jets to disperse gas into small bubbles so that the transfer of momentum, heat, and mass between phases can be achieved more efficiently. Various designs of two phase injectors-ejectors can be found in the literature (e.g., Jackson, 1964; Zlokarnik, 1979; Ogawa et al., 1983). Studies of fundamentals of jet reactors have been concentrated mainly on hydrodynamic properties, including gas holdup, bubble size, interfacial area, energy consumption, and mixing, and mass transfer properties of the reactors. No research has been done on heat transfer in the reactors. Furthermore, most studies have considered only the "lump", "global", or "overall" properties in the reactor system. Little research has been conducted to investigate "local" transport characteristics, specifically, those in the region of gas-liquid spouting. In a recent study of a jet reactor, Ogawa et al. (1983) concluded that the reactor can only be accurately accounted for by considering separately the behavior of both the spouting region and calming region. This
consideration is due to distinct differences in hydrodynamic and liquid phase mixing behavior and mass transfer characteristics between them.

The hydrodynamic properties in jet reactors due to gas-liquid spouting are discussed in the following:

1. Mixing

Using a stimulus-response technique and a diffusion model, Prokop et al. (1983) measured the axial dispersion coefficient of the liquid phase in the presence of electrolytes and ethanol. The liquid medium was derived from actual yeast cultivation. The effect of liquid spouting on the axial dispersion was found to be stronger than that of air spouting. For a 5 mm ID nozzle injector, the axial dispersion coefficient decreases with an increase in air velocity at low liquid velocities, and is independent of air velocity at high liquid velocities. Moreover, the axial dispersion coefficient is not constant throughout the entire reactor column due to circulation effects. They also found that the axial dispersion coefficient obtained in the jet reactor did not differ from those in bubble column without ejectors.
2. Gas Holdup

Weisweiler and Rosch (1978) reported that the gas holdup increases with gas velocity or the power of the jet. They correlated the overall gas holdup with the gas volumetric flow rate and the jet power per unit volume of reactor. Ogawa et al. (1983) also observed similar behavior in their work. However the empirical correlation equation they proposed are different from that of Weisweiler and Rosch (1978). They use gas velocity and Froude number as parameters to correlate the gas holdup in the spouting and calming regions.

3. Bubble Size and Interfacial Area

One distinct advantage of using two phase injectors-ejectors is that the bubble size is smaller than that with plate distributors. Heijnen and Van't Riet (1984) indicated that injectors-ejectors generate 0.5-1.0 mm bubbles in non-coalescing media and 4-6 mm bubble in coalescing media. Weisweiler and Rosch (1978) measured the bubble size in carbon dioxide/nitrogen-sodium hydroxide system by the photographic method and found that bubble size is in the range of 0.2-2 mm. They also calculated Sauter mean diameter and reported that the larger the energy input, the smaller the bubble mean diameter. Ogawa et al. (1983) observed that bubble size in the spouting region is the same as that in the calming region and proposed an empirical correlation equation for the bubble diameter.
Weisweiler and Rosch (1978) measured the total gas-liquid interfacial area by absorption of carbon dioxide in sodium hydroxide solution. They reported that the interfacial area increases with an increase in jet velocity or gas velocity. The interfacial area also increases with an increase in carbon dioxide concentration in the feed. They suggested that this is due to the fact that the "chemical" interfacial area is not the same as the total "geometric" interfacial area but represents only the fraction actually used for mass transfer. They also measured the interfacial area by the photographic method and compared the results with those obtained by the absorption method. The interfacial area obtained by the photographic method is about ten times as large as that obtained by the absorption method, due partially to gas backmixing. They suggested that further research needs to be done on gas backmixing, bubble size distribution, bubble residence time distribution and bubble coalescence to explain the difference in interfacial area obtained by the two methods. Ogawa et al. (1983) measured the gas-liquid interfacial area by absorption of oxygen to sodium sulfite solution and reported that the interfacial area in the spouting region is larger than that in the calming region.
C. Airlift Reactor

1. Gas Holdup

Gas holdup in airlift reactors has been subjected to extensive study (e.g., Hatch, 1973, Weiland and Onken, 1981a, 1981b; Fields and Slater 1983). Hatch (1973) and Fields and Slater (1983) reported that the gas holdup in internal loop airlift reactors increases with an increase in gas velocity. Weiland and Onken (1981a, 1981b) also observed similar behavior in external loop airlift reactors. The effect of the liquid velocity on the gas holdup is investigated by Weiland and Onken (1981b). They showed that the gas holdup decreases as the liquid velocity is increased. In the study of an external loop airlift reactor, Merchuk and Stein (1981) observed that the average gas holdup in the riser can be represented by the local gas holdup at the center of the riser.

2. Bubble Penetration Depth in the Annular Region

Fields and Slater (1983) observed in the study of an internal loop airlift reactor that a front formed which separates bubbly liquid from bubble free liquid in the annular region. This front moved downward as the gas flow rate was increased.
3. Liquid Circulation Rate

Hatch (1973) used hot water tracers to measure the liquid velocity in the draft tube region of an internal loop airlift reactor. They proposed a model to account for the liquid velocity in the draft tube region based on the pressure drops across the lengths of the draft tube and annulus and the pressure drop due to fluids turnaround at the base of the draft tube. Merchuk and Stein (1981) and Hsu and Dudukovic (1980) measured the liquid circulation rate in external loop airlift reactors. They also developed models to predict the liquid circulation rate using similar approach to that of Hatch (1973).

4. Bubble Size Distribution

Adler and Shugerl (1983) used an electrical conductivity microprobe to measure bubble size distribution in a single stage external loop airlift reactor. They reported that the mean and Sauter bubble diameters are constant and are independent of the operating conditions.

5. Solid-Liquid Mass Transfer

Extensive information of the solid-liquid mass transfer in liquid-solid fluidized beds has been available in the literature (e.g., Fan et al., 1960; Tournie et al., 1977; Ballesteros et al., 1982). However, little is known regarding the solid-liquid mass transfer in draft tube gas-liquid-solid spouted bed.
Several methods have been employed to obtain the solid-liquid mass transfer coefficient in fluidized beds: dissolution of a solute into the liquid phase (e.g., Rowe and Claxton, 1965; Ballesteros et al., 1982; Arters and Fan, 1984), adsorption (e.g., Gahno et al., 1975), crystallization (e.g., Carse and Mullin, 1968; Laguerie and Angelino, 1975), and ion exchange (e.g., Koloini et al., 1976; Rahman and Streat, 1981). By far, the first method is the simplest and most popular method.

The correlation of solid-liquid mass transfer coefficient is usually expressed in terms of some dimensionless numbers and the mass transfer coefficient is expressed in terms of Sherwood number or Colburn factor. The correlation equations proposed by Chu et al. (1953) and Fan et al. (1960) contains the bed void fraction. However, the bed void fraction is not involved in the correlation equations proposed by Tournie et al. (1979) and Ballesteros et al. (1982). Damronglerd et al. (1975) and Shen (1983) indicated that the bed void fraction is not a fundamental parameter of a system. It depends on the operating conditions of the system.
CHAPTER 3

HYDRODYNAMIC BEHAVIOR IN A DRAFT TUBE

GAS-LIQUID-SOLID SPOUTED BED

The hydrodynamics of a gas-liquid-solid fluidized bed with a draft tube is investigated. Specifically, the hydrodynamic properties under investigation include flow mode, pressure profile and pressure drop, bubble penetration depth in the annular region, overall gas holdup, apparent liquid circulation rate, and bubble size distribution in the draft tube and annular regions. The effects of gas and liquid velocities, solids loading, and physical properties of solid particles on these hydrodynamic properties are examined for two flow regimes in the draft tube region, namely, the dispersed bubble regime and the coalesced bubble regime.

EXPERIMENTAL

A schematic diagram of the experimental apparatus is shown in Fig. 3.1. The apparatus is made of a 15.2 cm ID plexiglass tube 122 cm in height. A 10.2 cm ID plexiglass draft tube 96 cm in height is coaxially situated inside the column. The bottom of the column is a
Figure 3.1 Schematic Diagram of the Experimental Apparatus.

1. Draft tube
2. Outside column
3. Injector
4. Particle withdrawing valve
5. Manometers
6. Upper retaining grid
7. Quick closing valves
8. Pressure gauge
9. Rotameters
10. Liquid pump
11. Liquid reservoir
conical section with a cone angle of 90 degrees. The gas and liquid injector is located at the bottom of the conical section. Details of the injector design are given in Fig. 3.2. The injector consists of two concentric tubes. The inside diameters of the outer and inner tubes are 2.54 cm and 1.27 cm, respectively. The gas enters the annulus of the injector through two 0.63 cm ID tubes located at 10.79 cm below the top of the injector. These two tubes are situated at opposite sides of the injector. The gas then enters the column through a perforated plate at the top of the injector. There are sixteen 0.12 cm ID holes on the perforated plate. The liquid enters the column from the 1.27 cm ID tube. A ball valve below the injector is utilized to withdraw solid particles from the column. A movable upper retaining grid, which serves to constrain the bed, is placed at the top of the column. In this study, the upper retaining grid is fixed at a location immediately underneath the fluid outlet. Thus, the total length of the bed, from the top of the injector to the upper retaining grid, is 120.6 cm. Four types of solid particles are used in the experiments. They are 3 mm glass beads, 4 mm glass beads, 2.2 mm alumina particles, and 2.5 mm nylon balls. The physical properties of these particles are given in Table 3.1.

Water and air are used as the liquid and gas phases in the experiments. The superficial liquid and gas velocities in the experiments vary from 0 to 6 cm/sec and from 0 to 5 cm/sec, respectively, based on the cross-sectional area of the outside column. A quick closing valve is employed to allow simultaneous shut-off of
Figure 3.2 Schematic Diagram of the Gas-Liquid Injector
Table 3.1 Physical Properties of the Particles Used in the Experiments

<table>
<thead>
<tr>
<th>Type</th>
<th>Material</th>
<th>Diameter (mm)</th>
<th>Density (g/cc)</th>
</tr>
</thead>
<tbody>
<tr>
<td>I</td>
<td>Glass Beads</td>
<td>3.0</td>
<td>2.52</td>
</tr>
<tr>
<td>II</td>
<td>Glass Beads</td>
<td>4.0</td>
<td>2.47</td>
</tr>
<tr>
<td>III</td>
<td>Alumina</td>
<td>2.2</td>
<td>3.73</td>
</tr>
<tr>
<td>IV</td>
<td>Nylon Balls</td>
<td>2.5</td>
<td>1.15</td>
</tr>
</tbody>
</table>
both the gas and liquid flows when measurement of the overall holdups of gas and liquid are desired. There are thirteen pressure taps placed along the outside column and nine along the draft tube. Water manometers are used to acquire the pressure distribution data for both the outside column and the draft tube. To obtain the apparent liquid circulation rate between the draft tube and the annulus, the liquid circulation time is measured using a solid tracer particle made of polystyrene. The tracer particle is 5 mm in diameter with a specific gravity of 1.003. The circulation time, \( \bar{t} \), for the tracer particle to make a single loop from the draft tube region to the annular region and back is related to the apparent liquid circulation velocity, \( U_C \), by

\[
U_C = \frac{4V_L}{\bar{t} \pi D_c^2}
\]  

(1)

where \( V_L \) is the volume of the bed occupied by the liquid phase. In this study \( \bar{t} \) averaged over fifty observations is used for each calculation of \( U_C \).

The local gas holdup and bubble size distribution are obtained by using a twin-electrode conductivity probe (Matsuura and Fan, 1983). The probe consists of two stainless steel syringe needles with the diameter of the tip of syringe needle exposed to the environment of 0.2 mm. The needles are separated by epoxy resin. The vertical distance between the two needles is 0.3 mm. The probe is supported by a stainless steel tube which serves as an opposite electrode. Details of the probe signal processing unit, signal acquisition and matching
of a single pair, and data analysis are given in Matsuura and Fan (1983).

RESULTS AND DISCUSSION

The results of the hydrodynamic properties study including flow mode, pressure profile and pressure drop, bubble penetration depth in the annular region, overall gas holdup, apparent liquid circulation rate, and bubble size distribution in the draft tube and annular regions for a draft tube gas-liquid-solid spouted bed are presented and discussed in the following:

A. Flow Mode

The flow mode of the system with liquid as the continuous phase can be distinguished by visual observation for the various operating conditions. Three distinct modes are observed. They are the packed bed mode, the fluidized bed mode, and the circulated bed mode. Flow mode maps for selected experimental conditions are shown in Figs. 3.3, 3.4, and 3.5. At low gas and liquid velocities, the bed is in the packed bed mode where all the particles are packed at the bottom of the draft-tube and annular regions of the column. As the gas or liquid velocity increases, the solid particles under the draft tube become fluidized to form a spout. The onset of the fluidized bed mode is defined as the point where the global downward movement of the solid particles occurs around the wall area of the conical section of
Figure 3.3 Flow Mode Map for Various Types of Particles
Figure 3.4 Flow Mode Map for 3 mm Glass Beads for Various Solids Loading
Figure 3.5 Flow Mode Map for 2.5 mm Nylon Balls for Various Solids Loading
the bed. With further increase in the gas or liquid velocity beyond the onset point, the solid particles under the annular region begin to massively migrate to the draft tube region to form a fluidized bed. In the fluidized bed mode, considerable liquid circulation occurs and some bubbles are circulated into the annular region. As the gas or liquid velocity is further increased, the height of the fluidized bed increases. When the fluidized bed height exceeds the draft tube height, the solid particles are carried over into the annular region. These particles are then circulated back to the draft tube. This flow mode characterizes the circulated bed mode. In this flow mode, a large number of bubbles are observed in the annular region. An intensive solid particle circulation is also evident.

Among these three flow modes, the circulated bed mode is of particular importance when an application of the draft tube spouted bed to biological processing is desired. This is due to the fact that the circulated bed mode provides high gas holdup and high contacting efficiency between the gas phase and the liquid phase and the solid phase.

Also shown in Fig. 3.3 is the effect of particle size and density on the flow mode. It is seen that for a larger particle diameter at a constant particle density, the transition from the packed bed mode to the fluidized bed mode or from the fluidized bed mode to the circulated bed mode occurs at a higher gas velocity for a constant liquid velocity or at a higher liquid velocity for a constant gas velocity. It is noted that the curves shown for 4 mm glass beads and
2.2 mm alumina particles are similar due to the fact that the terminal velocities for these particles in water are about the same (42.3 cm/sec for 4 mm glass beads, and 42.7 cm/sec for 2.2 mm alumina particles).

The effect of the solids loading on the flow modes is shown in Figs. 3.4 and 3.5 for 3 mm glass beads and 2.5 mm nylon balls, respectively. It is seen that it requires a higher liquid or gas velocity to reach the circulated bed mode from the fluidized bed mode for a high solids loading than for a low solids loading. This appears to contradict the common notion derived from the conventional gas-liquid-solid fluidized bed in which for given gas and liquid velocities, higher solids loading would yield higher bed expansion. Consequently, to achieve a given height of bed expansion for a conventional fluidized bed, a low solids loading in the bed requires a higher gas or liquid velocity than does a high solids loading in the bed. It should be noted, however, that in the draft tube spouted bed, a considerable amount of the circulated liquid contributes to the liquid flow in the draft tube region. Furthermore, the bed expansion in the draft tube region of the spouted bed is directly dictated by the liquid flow rate and gas holdup. As will be shown later, when the solids loading increases, the liquid circulation rate between the draft tube region and the annular region first increases and then decreases and the overall gas holdup monotonically decreases. This coupling behavior of the liquid circulation rate and the overall gas holdup over the increase of solids loading are responsible for the
flow mode map given in the figure. In what is to follow, the results for the 3 mm glass beads with a density of 2.52 g/cc and the 2.5 mm nylon balls with a density of 1.15 g/cc are reported.

B. Pressure Distribution

Typical pressure distributions in both the draft tube region and the annular region for 3 mm glass beads and 2.5 mm nylon balls are shown in Figs. 3.6 and 3.7, respectively. At the level corresponding to the top of the draft tube, the pressure in the draft tube region is higher than that in the annular region. However, at the level corresponding to the bottom of the draft tube, the pressure in the draft tube region is lower than that in the annular region. This pressure difference provides adequate driving force for liquid circulation from the draft tube region to the annular region. It is seen in Fig. 3.6 that for the 3 mm glass beads, the pressure variation is approximately linear with the axial distance from the bottom to the top of the draft tube in the annular region. Clearly, in this region the pressure variation can be estimated directly from the weight of each phase holdup. This is not the case, however, for the pressure distribution in the draft tube region. The drastic change in the pressure profile near the bottom of the draft tube in the draft tube region as shown in the figure could be attributed to the liquid and gas jetting effects which are strongly dependent on the physical properties of the particles used. It is seen in Fig. 3.7 that for the 2.5 mm nylon balls, the pressure varies linearly with the axial
Figure 3.6 Pressure Distribution in the Draft Tube Region and the Annular Region for 3 mm Glass Beads. L is the Downward Distance from the Upper Retaining Grid.
Figure 3.7 Pressure Distribution in the Draft Tube Region and the Annular Region for 2.5 mm Nylon Balls. L is the Downward Distance from the Upper Retaining Grid.
distance in both the draft tube and the annular regions. This pressure variation behavior signifies a minimum fluid jetting effect at the draft tube entrance. More extensive analysis is necessary to understand the jetting phenomenon in the draft tube region of the system.

The following mathematical model is proposed to describe the pressure drop in the bed:

Consider a bed consisting of four regions; namely, below-draft-tube region, draft tube region, above-draft-tube region, and annular region as shown in Fig. 3.8. The momentum balance for each of the four regions gives rise to the following equations:

(i) Below-Draft-Tube Region

\[
(r_{g_1} g_{11} g_{1} + \rho_{s_1} u_{s_1}^2 A_{s_1}) - (r_{g_2} g_{22} g_{2} + \rho_{s_2} u_{s_2}^2 A_{s_2}) - (r_{g_3} g_{33} g_{3} + \rho_{s_3} u_{s_3}^2 A_{s_3})
\]

\[
+ \rho_{s_3} u_{s_3}^2 A_{s_3} + (P_{A_1} - P_{A_2} - P_{A_3})
\]

\[- (r_{g_5} g_{55} + \rho_{s_5} u_{s_5}^2 V_{B} - F_B = 0]

(ii) Draft-Tube Region

\[
(r_{g_3} u_{g_3}^2 g_{3} + \rho_{s_3} u_{s_3}^2 A_{s_3} + \rho_{s_3} u_{s_3}^2 A_{s_3}) - (r_{g_4} u_{g_4}^2 g_{4})
\]

\[
+ \rho_{s_4} u_{s_4}^2 A_{s_4} + \rho_{s_4} u_{s_4}^2 A_{s_4} + (P_{3_{A_3}} - P_{4_{A_4}})
\]

\[- (r_{g_5} g_{55} + \rho_{s_5} u_{s_5}^2 V_{D} - F_D = 0]

Figure 3.8 Schematic Representation of the Bed for the Description of the Pressure Drop Behavior and the Apparent Liquid Circulation Rate
(iii) Above-Draft-Tube Region

\[ (\rho \frac{U^2}{2} A_4 + \rho_2 \frac{U^2}{2} A_6 + \rho_4 \frac{U^2}{2} A_4) + (\rho_5 \frac{U^2}{2} A_5 + \rho_5 \frac{U^2}{2} A_5) \]
\[ + \rho_5 \frac{U^2}{2} (A_5) - (\rho_6 \frac{U^2}{2} A_6 + \rho_6 \frac{U^2}{2} A_6) \]
\[ + (\rho \frac{p^2}{2} A_4 + \rho_5 \frac{p^2}{2} A_5 - \rho_6 \frac{p^2}{2} A_6) - (\rho \frac{\epsilon}{2} T + \rho \frac{\epsilon}{2} T + \rho \frac{\epsilon}{2} T)V_T g \]
\[ - F_T = 0 \]

(iv) Annular Region

\[ -(\rho \frac{U^2}{2} A_4 + \rho_2 \frac{U^2}{2} A_6 + \rho_4 \frac{U^2}{2} A_4) + (\rho_5 \frac{U^2}{2} A_5 + \rho_5 \frac{U^2}{2} A_5) \]
\[ + \rho_2 \frac{U^2}{2} A_2 + (\rho_5 A_5 - \rho_6 A_6) - (\rho \frac{\epsilon}{2} A_6 + \rho \frac{\epsilon}{2} A_6 + \rho \frac{\epsilon}{2} A_6) V_A g + F_A = 0 \]

Adding Eqns. (2) through (4) to Eqn. (5), yields

\[ (\rho \frac{U^2}{2} A_1 + \rho_1 \frac{U^2}{2} A_2) - (\rho \frac{U^2}{2} A_6 + \rho \frac{U^2}{2} A_6) + (\rho \frac{U^2}{2} A_1) \]
\[ - (\rho \frac{\epsilon}{2} + \rho \frac{\epsilon}{2} + \rho \frac{\epsilon}{2}) V_A g - F = 0 \]

where

\[ \rho \frac{\epsilon}{2} + \rho \frac{\epsilon}{2} + \rho \frac{\epsilon}{2} = (\rho \frac{\epsilon}{2} A + \rho \frac{\epsilon}{2} A + \rho \frac{\epsilon}{2} A) \frac{V_A}{V_A} \]
\[ + (\rho \frac{\epsilon}{2} B + \rho \frac{\epsilon}{2} B + \rho \frac{\epsilon}{2} B) \frac{V_B}{V_A} + (\rho \frac{\epsilon}{2} D + \rho \frac{\epsilon}{2} D + \rho \frac{\epsilon}{2} D) \frac{V_D}{V_A} \]
\[ + (\rho \frac{\epsilon}{2} T + \rho \frac{\epsilon}{2} T + \rho \frac{\epsilon}{2} T) \frac{V_T}{V_A} \]
\[ F = (F_b + F_d + F_T - F_A) \]  

(8)

\[ F \] can also be expressed as

\[ F = (\Delta P_f) \left( \frac{\pi}{4} D_c^2 \right) \]  

(9)

where \( \Delta P_f \) is the frictional pressure loss. It should be noted that the experimental results indicate that the pressure drop across the upper retaining grid is negligibly small compared to the frictional pressure drop in the bed. The friction factor, \( f \), can be defined based on the modified Fanning equation as given by

\[ \Delta P_f = 4f \left( \frac{L_1}{D_c} \right) \left( \frac{1}{2} \right) (\rho_m U_c^2) \]  

(10)

where \( L_1 \) is the distance from the top of the distributor to the upper retaining grid, \( D_c \) is column diameter, \( \rho_m \) is mixture density defined as \( (\rho_g \varepsilon_g + \rho_L \varepsilon_L + \rho_s \varepsilon_s) \), and \( U_c \) is the apparent liquid circulation velocity based on the cross-sectional area of the column. Based on the experimental data obtained in this study, the value of \( f \) is found to be a unique function of \( (Re)_c \) as shown in Fig. 3.9, where \( (Re)_c \) is the apparent liquid circulation Reynolds number defined by \( D_c U_c \rho_L / \mu_L \). It is seen that \( f \) varies linearly on a log-log scale with respect to \( (Re)_c \). Furthermore, \( f \) decreases with the increase of \( (Re)_c \) according to the following empirical equation:

\[ f = 1.28 \times 10^9 (Re)_c^{-1.74} \]  

(11)
Figure 3.9 Variation of the Friction Factor, $f$, with $(Re)_c$, where $(Re)_c$ is Defined as $D_c U_c \rho / \mu _c$
Figure 3.10 Comparison of the Predicted and Experimental Values for the Pressure Drop in the Bed
The pressure drop in the draft-tube spouted bed calculated by the model equations, i.e., Eqns. (6), (9), (10), and (11), is compared with that obtained experimentally as shown in Fig. 3.10. It is seen that this comparison is also satisfactory except for three data points. The three data points represent data from the highest liquid velocity (6 cm/sec) at three different gas velocities. Such a high liquid velocity results in excessively high apparent liquid circulation rate and hence, high frictional pressure drop.

C. Bubble Penetration Depth in the Annular Region

Fields and Slater (1983) observed in their study of an internal loop airlift reactor that a front formed which separates bubbly liquid from bubble free liquid in the annular region (downcomer). In the present study of a draft tube gas-liquid-solid spouted bed, a similar bubbly liquid front was also observed. However, this front does not appear to be stable under all operating conditions considered in this study; it fluctuates to a large degree. Figures 3.11 and 3.12 show the effect of the gas and liquid velocities on the bubble penetration depth in the annular region for gas velocities from 1 to 5 cm/sec and liquid velocities from 0.5 to 4.5 cm/sec for 3 mm glass beads. Figure 3.13 shows the effect of the gas and liquid velocities on the bubble penetration depth in the annular region for gas velocities from 0.2 to 0.8 cm/sec and liquid velocities from 0.52 to 2.11 cm/sec for 2.5 mm nylon balls. In these figures, the bubble penetration depth is defined as the distance from the top of the draft tube to the bubbly
Figure 3.11 Effect of the Gas Velocity on the Bubble Penetration Depth in the Annular Region as a Function of the Liquid Velocity for 3 mm Glass Beads. $H$ is the Downward Distance from the Top of the Draft Tube
Figure 3.12 Effect of the Liquid Velocity on the Bubble Penetration Depth in the Annular Region as a Function of the Gas Velocity for 3 mm Glass Beads. $H$ is the Downward Distance from the Top of the Draft Tube.
Figure 3.13 Effect of the Gas Velocity on the Bubble Penetration Depth in the Annular Region as a Function of the Liquid Velocity for 2.5 mm Nylon Balls. H is the Downward Distance from the Top of the Draft Tube.
liquid front, which is taken as an arithmetic mean of the maximum and minimum penetration depths at a given operating condition. As shown in Figs. 3.11 and 3.13, the bubble penetration depth increases as the gas velocity is increased, which follows the same behavior as that reported by Fields and Slater (1983). As the gas velocity increases, a critical gas velocity is reached under which the bubbly liquid front reaches the level corresponding to the bottom of the draft tube. With any further increase of the gas velocity beyond the critical gas velocity, complete bubble circulation is achieved. It is noted from Figs. 3.11 and 3.13 that the critical gas velocity decreases with an increase in liquid velocity. The bubble penetration depth increases as the liquid velocity increases as shown in Fig. 3.12.

It should be noted that the behavior of the bubble penetration in the annular region resembles that of bed expansion in inverse gas-liquid-solid fluidization (Fan et al., 1982a, 1982b; Chern et al., 1983). For a bubble to be entrained over infinite depth, the liquid circulation velocity would be that required to exceed the terminal velocity of a bubble. Thus, at low liquid circulation velocity, only a small number of bubbles are entrained to the annular region. As the liquid circulation velocity increases, more bubbles are entrained to the annular region which consequently enhances the bubble penetration depth in the annular region. Figures 3.14 and 3.15 show the effect of solids loading on the bubble penetration depth for 3 mm glass beads and 2.5 mm nylon balls, respectively. It is seen that the penetration depth decreases and the
Figure 3.14  Effect of the Solids Loading on the Bubble Penetration Depth in the Annular Region for 3 mm Glass Beads. $H$ is the Downward Distance from the Top of the Draft Tube

3 mm glass beads
$U_I = 3.41$ cm/sec

$V_S$ (ml)

- 150 $\triangle$
- 300 $\square$
- 600 $\circ$

Bottom of the Draft Tube
Figure 3.15 Effect of the Solids Loading on the Bubble Penetration Depth in the Annular Region for 2.5 mm Nylon Balls. H is the Downward Distance from the Top of the Draft Tube
critical gas velocity increases with an increase in solids loading. Note that it was observed in the experiments that the concentration (number) of bubbles within the depth of penetration increases considerably when the solids loading increases. Rapid bubble coalescence and redispersion were also observed to periodically occur around the bubbly liquid front for the high solids loading condition.

D. Overall Gas Holdup

Figures 3.16 and 3.17 show the effect of the gas velocity on the overall gas holdup as a function of the liquid velocity for 3 mm glass beads and 2.5 mm nylon balls, respectively. The overall gas holdup increases as the gas velocity increases. Similar behavior was also observed in internal loop airlift reactors (e.g., Hatch, 1973; Fields and Slater, 1983) and external loop airlift reactors (e.g., Weiland and Onken, 1981a, 1981b). Note that the overall gas holdup and gas velocity follow a linear relationship which can be extrapolated to the origin. This signifies that the relationship satisfies the physical constraint in which at zero gas velocity the overall gas holdup is zero. Comparisons of the overall gas holdup between the draft tube gas-liquid-solid spouted bed and the conventional gas-liquid-solid fluidized bed are also made in the figures. The overall gas holdup of the conventional gas-liquid-solid fluidized bed operated in the similar bed and particle size ranges to the present system can be estimated by the empirical correlation given by (Begovich and Watson, 1978):
Figure 3.16 Effect of the Gas Velocity on the Gas Holdup as a Function of the Liquid Velocity for 3 mm Glass Beads

3mm glass beads
$V_S = 300\ ml$
parameter: $U_j,\ cm/sec$

Figure 3.17 Effect of the Gas Velocity on the Gas Holdup as a Function of the Liquid Velocity for 2.5 mm Nylon Balls
\[ \varepsilon_g = 0.048 u_g^{0.72} d_p^{0.168} D_c^{-0.125} \]  

(12)

Note that the overall gas holdup expressed by this equation is independent of the liquid velocity. This is due to the fact that in a conventional gas-liquid-solid fluidized bed, for a given gas velocity, both the bed height and the bubble frequency increase with an increase in the liquid velocity. As a consequence, the overall gas holdup remains unchanged. As shown in the figures, the overall gas holdup of the draft tube gas-liquid-solid spouted bed is substantially higher than that of the conventional gas-liquid-solid system, particularly at the high gas or liquid flow condition.

With additional data beyond those shown in Fig. 3.16, effects of the liquid velocity on the overall gas holdup are distinctly shown in Fig. 3.18. It is seen that the overall gas holdup increases with increasing liquid velocity. This behavior is in contrast to that observed for the external loop reactor by Weiland and Onken (1981b), in which the overall gas holdup decreases with increasing liquid velocity. Note that in the external loop reactor, gas is separated from the reactor at the top of the column (riser), and thus, no gas bubbles are present in the downcomer of the reactor. As indicated earlier, in the present system there is considerable bubble entrainment from the draft tube region to the annular region by the circulated liquid. This contributes to the large overall gas holdup.
Figure 3.18 Effect of the Liquid Velocity on the Gas Holdup as a Function of the Gas Velocity for 3 mm Glass Beads
in the system. This important behavior allows the draft tube gas-liquid-solid spouted bed to be operated at high liquid throughputs while maintaining high contacting efficiency between the gas and liquid phases. Figures 3.19 and 3.20 show the effect of solids loading on the overall gas holdup for 3 mm glass beads and 2.5 mm nylon balls, respectively. It is seen that the overall gas holdup decreases with the increase of the solid loading. For all the solids loadings considered, the overall gas holdup of the draft tube gas-liquid-solid system is higher than that of the conventional gas-liquid-solid fluidized bed. An empirical correlation to account for the overall gas holdup in the draft tube gas-liquid-solid spouted bed has been obtained. It was found that the overall gas holdup can be related to \( (Re)_C \) by:

\[
\varepsilon_g = -0.115 + 1.045 \times 10^{-5} (Re)_C
\]  

(13)

Figure 3.21 shows the experimental data and correlation equation for \( \varepsilon_g \) versus \( (Re)_C \) with solids loading varying from 0 to 600 ml. It is noted that the correlation coefficient for Eqn. (13) is 0.85.

E. Apparent Liquid Circulation Rate

In a strict sense, the liquid circulation rate in the draft tube gas-liquid-solid spouted bed is represented by the liquid flow rate observed in the annular region. However, the liquid flow rate in the annular region is so large that the residence time of the tracer particle in this region is on the order of one to two seconds.
Figure 3.19 Effect of the Solids Loading on the Gas Holdup for 3 mm Glass Beads
Figure 3.20 Effect of the Solids Loading on the Gas Holdup for 2.5 mm Nylon Balls
Figure 3.21 Variation of the Overall Gas Holdup with \( (Re)_c \), where \( (Re)_c \) is Defined as \( D_c \frac{U_c}{\nu_L} \).
Consequently, using the present solid tracer technique to evaluate the liquid circulation rate cannot lead to an accurate result on liquid flow rate in the annular region. To characterize the liquid circulation rate, however, the "apparent" liquid circulation rate, $Q_l$, as defined in this study is given by

$$Q_l = \frac{V_l}{\bar{\epsilon}}$$ (14)

As was shown earlier, $Q_l$ or $U_c (4Q_l/\pi D_c^2)$ is an effective parameter for use in the correlation of various hydrodynamic properties of the system, e.g., pressure drop and overall gas holdup. Moreover, the present study shows that $Q_l$ does reflects to a great extent the "true" liquid circulation rate.

The effects of the gas and liquid velocities on the apparent liquid circulation rate for 3 mm glass beads are shown in Figs. 3.22 and 3.23, respectively. It is seen that the apparent liquid circulation rate increases with an increase in gas velocity over a wide range of gas velocities. In Fig. 3.22, the apparent liquid circulation rate is also found to increase with an increase in liquid velocity over a wide range of liquid velocities. As shown in Fig. 3.23, when the liquid velocity is less than 2.25 cm/sec, which corresponds to the coalesced bubble flow regime in the orifice tube region, the apparent liquid circulation rate increases drastically as the liquid velocity is increased. However, when the liquid velocity is increased beyond 2.25 cm/sec, which corresponds to the dispersed bubble flow regime in the draft tube region, the apparent liquid
Figure 3.22 Effect of the Gas Velocity on the Apparent Liquid Circulation Rate as a Function of the Liquid Velocity for 3 mm Glass Beads

3mm glass beads
parameter: $U_1$, cm/sec
$V_5 = 300$ ml

$Q_l$, l/sec
$U_g$, cm/sec
Figure 3.23 Effect of the Liquid Velocity on the Apparent Liquid Circulation Rate as a Function of the Gas Velocity for 3 mm Glass Beads
circulation rate first increases slowly with increasing liquid velocity until a liquid velocity of 5.2 cm/sec is reached. Beyond this velocity, the apparent liquid circulation rate increases dramatically with liquid velocity. Also shown in the figure is the apparent liquid circulation rate for the airlift system. It is seen that in the dispersed bubble regime, the airlift system gives rise to a lower apparent liquid circulation rate than the draft tube gas-liquid-solid spouted bed. In the coalesced bubble regime, however, no such conclusion can be reached. Figure 3.24 shows the effect of solids loading on the apparent liquid circulation rate for 3 mm glass beads for a liquid velocity of 3.41 cm/sec. It is seen that at a gas velocity beyond 2 cm/sec, as the solids loading increases, the apparent liquid circulation rate first increases then decreases. Apparently, under this gas velocity range, an optimal solids loading exists which gives rise to a maximum apparent liquid circulation rate. It is seen also that the apparent liquid circulation rate for an airlift system is lower than that of the draft tube gas-liquid-solid spouted bed at a gas velocity beyond 3.6 cm/sec when the solids loading is less than 600 ml.

A mathematical model is proposed in this study to account for the apparent liquid circulation rate in the draft tube gas-liquid-solid spouted bed. The model can be described by the following:
Figure 3.24 Effect of the Solids Loading on the Apparent Liquid Circulation Rate for 3 mm Glass Beads
Referring to Fig. 3.8, the pressure drop in the draft tube region, i.e., \( P_3 - P_4 \), can be given by
\[
P_3 - P_4 = \Delta P_{\text{Df}} + \Delta P_{\text{Da}} + \Delta P_{\text{Dh}} \tag{15}
\]
where \( \Delta P_{\text{Df}} \) = pressure drop in the draft tube region due to friction
\( \Delta P_{\text{Da}} \) = pressure drop in the draft tube region due to fluid acceleration
\( \Delta P_{\text{Dh}} \) = pressure drop in the draft tube region due to hydrostatic head

Neglecting the pressure drop due to friction and acceleration, Eqn. (15) can be simplified to
\[
P_3 - P_4 = \Delta P_{\text{Dh}} \tag{16}
\]
Similarly, the pressure drop in the annular region is given as
\[
P_2 - P_5 = \Delta P_{\text{Ah}} \tag{17}
\]
where \( \Delta P_{\text{Ah}} \) is the pressure drop in the annular region due to hydrostatic head. Subtracting Eqn. (16) from Eqn. (17), yields
\[
(P_4 - P_5) + (P_2 - P_3) = \Delta P_{\text{Ah}} - \Delta P_{\text{Dh}} \tag{18}
\]
The right hand side of Eqn. (18) can be expressed in terms of the bulk density difference in the annular region and the draft tube region as follows
\[
(\Delta P_{\text{Ah}} - \Delta P_{\text{Dh}}) = (\rho g \varepsilon gA + \rho_s \varepsilon sA) g L_D
- (\rho g \varepsilon gD + \rho_s \varepsilon sD) g L_D \tag{19}
\]
where \( L_D \) is the height of the draft tube. The pressure loss due to
fluid circulation from point 4 to point 5 defined in the figure can be decomposed into two components contributing to this pressure loss: fluid expansion from the draft tube region to the above-draft-tube region and fluid contraction from the above-draft-tube region to the annular region. Accordingly, $P_4 - P_5$ can be expressed by

$$
(p_4 - p_5) = \frac{1}{2} \rho \frac{U_D}{A_D}^2 (1 - A_D/A_T)^2 + \frac{1}{2} K_{\rho} \frac{U_A}{A_T}^2
$$

(20)

where $U_D$ = the liquid velocity in the draft-tube region based on the draft tube area

$U_A$ = the liquid velocity in the annular region based on the annular region

$A_T$ and $A_D$ = the cross-sectional areas of column and draft tube regions, respectively

$K$ = the sudden contraction loss coefficient

$K$ in Eqn. (20) is a function of $A_T/A_D$. The value of $K$ can be approximated by that for single phase liquid flow (Perry and Chilton, 1973) due to the fact that the system is predominately occupied by the continuous liquid phase.

The pressure loss due to fluid circulation from point 2 to point 3 defined in Fig. 3.8 can also be decomposed into two components contributing to this pressure loss: gradual contraction from the annular region to the injector and liquid jetting from the injector to the draft tube region. Accordingly, $P_2 - P_3$ can be expressed as

$$
(p_2 - p_3) = \frac{1}{2} \zeta \rho \frac{U_B}{A_B}^2 + \frac{1}{2} \rho \left(\frac{U_1}{A_T}^2 - \frac{U_D}{A_D}^2\right)
$$

(21)

where $U_1$ = the liquid velocity from the injector

$U_B$ = the liquid velocity in the annular region based on the cross-sectional area of the mid-outer tube
of the injector; i.e., area with radius of $r_i + 1/2(r_i - r_o)$, where $r_i$ and $r_o$ are the radii of the inner and outer tubes of the injector, respectively.

$\zeta = \text{the gradual contraction loss coefficient}$

Note that for the second term on the right hand side of Eqn. (21), the kinetic energy loss is considered to dominate the pressure loss due to liquid jetting from the injector to the draft tube region. $\zeta$ in Eqn. (21) can be related to the angle of the conical section, $\alpha$, by (Zhou, 1979)

$$\zeta = \frac{4f}{8\sin(\alpha/2)} \left(1 - 1/n^2\right) \quad (22)$$

where $\alpha = \text{the angle of the conical section}$

$n = \text{area ratio of the annular region to the mid-outer tube of the injector}$

$f = \text{the Fanning friction factor which can be evaluated by the following Blasius equation (Blasius, 1913)}$

$$f = 0.0791 \left(\text{Re}_A\right)^{-0.25} \quad (23)$$

with $(\text{Re}_A)$ defined by $D_e U_p \mu / \mu_A$, where $D_e$ is the equivalent diameter of the annular region defined by $4n^2$, with $n^2$ being the hydraulic radius. Note that the cross-sectional area of the injector is very small compared to that of the annular region, therefore the term $(1 - 1/n^2)$ in Eqn. (22) is almost equal to 1. $f$ characterizes the flow condition prior to entering the conical contraction section. The use of equivalent diameter expressed in terms of the hydraulic radius is a reasonable way to represent the present draft tube system.
Material balance for the liquid phase at the below-draft-tube and above-draft-tube regions gives rise to the following equations

\[ U_{D,A} = U_{A,A} + U_{A,L} \]  \hspace{1cm} (24)

\[ U_{D,A} = U_{A,A} + U_{B,B} \]  \hspace{1cm} (25)

The liquid velocity in the draft tube region, \( U_D \), and in the annular region, \( U_A \), can be obtained by solving Eqns. (18) through (25). The liquid circulation time in each of the four regions can be calculated by

\[ t_D = \frac{L_D \varepsilon_D}{U_D} \]  \hspace{1cm} (26)

\[ t_A = \frac{L_D \varepsilon_A}{U_A} \]  \hspace{1cm} (27)

\[ t_B = \frac{V_B \varepsilon_B}{U_{D,A}} \]  \hspace{1cm} (28)

\[ t_T = \frac{V_T \varepsilon_T}{U_{D,A}} \]  \hspace{1cm} (29)

The total liquid circulation time is the sum of the liquid circulation time in each of the four regions:

\[ t_{Total} = t_A + t_B + t_D + t_T \]  \hspace{1cm} (30)

Note that in calculating \( t_A \), \( t_B \), \( t_D \), and \( t_T \) from Eqns. (26) through (29), the experimental values for the liquid holdup in each of the four regions; i.e., \( \varepsilon_A \), \( \varepsilon_B \), \( \varepsilon_D \), and \( \varepsilon_T \) are used. The liquid holdup in each region is represented by the liquid holdup obtained at
the center of each region. This representation has been verified by Merchuk and Stein (1981) for the airlift system. As indicated earlier, the point gas holdup is determined by a twin-electrode conductivity probe (Matsuura and Fan, 1983) in this study. The liquid holdup can then be obtained from the pressure drop data for each region and the relationship that \( \varepsilon_{g} + \varepsilon_{l} + \varepsilon_{s} \) is equal to 1 where \( i = A, B, D \) and \( T \). The apparent liquid circulation rate can then be calculated by

\[
Q_{l} = \frac{V_{l}}{t_{\text{Total}}} \quad (31)
\]

Figure 3.25 shows the comparison of the apparent liquid circulation rate calculated by Eqn. (31) and the apparent liquid circulation rate obtained experimentally. It is seen that the comparison is satisfactory.

F. Bubble Size Distribution

Figure 3.26 shows the bubble size distributions for the draft tube and annular regions for the dispersed bubble regime in the draft tube region. Note that \( \xi \) and \( R(\xi) \) shown in this figure are dimensionless bubble size and bubble size distribution function, respectively. The bubble size distribution is defined based on the number frequency of the bubbles. It is seen that the bubble size distributions in the draft tube and annular regions follows log-normal
Figure 3.25 Comparison of the Predicted and Experimental Values for the Apparent Liquid Circulation Rate
Figure 3.26 Bubble Size Distributions Based on Number of Bubbles at the Center Points of the Draft Tube and Annulus for the Dispersed Bubble Regime
Figure 3.27 Bubble Size Distributions Based on Number of Bubbles at the Center Points of the Draft Tube and Annulus for the Coalesced Bubble Regime.
distribution function. The range of the bubble size distribution in the draft tube region is slightly larger than that in the annular region where the largest bubble observed is about 1.1 cm. The average bubble size is found to be 0.6 cm for the draft tube region and 0.38 cm for the annular region. The average bubble size for the distribution based on the volume frequency of the bubbles defined by (Matsuura and Fan, 1983)

$$\bar{d}_v = \frac{\int_0^\infty R(d_b) d_b^4 \, d d_b}{\int_0^\infty R(d_b) d_b^3 \, d d_b}$$

is found to be 1.75 cm for the draft tube region and 1.26 cm for the annular region. $R(d_b)$ in Eqn. (32) is the dimensional bubble size distribution function. Note that Eqn. (32) is defined by assuming a spherical bubble geometry. Figure 3.27 shows the bubble size distributions in the draft tube and annular regions for the coalesced bubble regime. Again, the bubble size distributions follow the log-normal distribution function. The average bubble size and size distribution in the annular region is similar to that observed in the same region for the dispersed bubble regime. Considerably larger bubble sizes exist, however, in the draft tube region compared to that in the same region for the dispersed bubble regime. The average bubble size for the distribution based on the volume frequency of bubbles is 2.23 cm for the draft tube region and 0.91 cm for the annular region.
CHAPTER 4

SOME DESIGN CONSIDERATIONS IN A DRAFT TUBE
GAS- LIQUID-SOLID SPouted BED

One of the most important dependent variables reflecting the performance of a draft tube gas-liquid-solid spouted bed is gas-liquid mass transfer. Fan et al. (1984c) showed that the gas-liquid mass transfer coefficient in a gas-liquid-solid spouted bed increases with an increase in the overall gas holdup for the overall gas holdup between 0 and 0.14. In a draft tube gas-liquid system, Koide et al. (1983) also reported that for the overall gas holdup between 0.037 and 0.21, the gas-liquid mass transfer coefficient varies linearly on a logarithmic scale with the overall gas holdup. Thus, effects of the design variables on the overall gas holdup qualitatively portray the effects on the gas-liquid mass transfer coefficient. The overall gas holdup is thus chosen as a convenient dependent variable on which the optimal design of a draft tube gas-liquid-solid spouted bed is based.

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The schematic diagram of the experimental apparatus is shown in Fig. 3.1 in Chapter 3. Draft tubes with inside diameters of 7.62 cm, 10.16 cm, and 12.7 cm are used in the experiments. Four types of injectors are employed, i.e., injectors 1, 2, 3, and 4 as shown in Fig. 4.1. Injector 1 is the same as that shown in Fig. 3.2 in Chapter 3. The design of injectors 2 and 3 are the same as that of injector 1 except that the perforated plate is replaced by porous plates of HDPE (high density polyethylene) and PMMA (polymethylmethacrylate), respectively. The average pore size of HDPE and PMMA porous plates are 15 μm and 30 μm, respectively. Injector 4 consists of two concentric tubes. The inside diameters of the outer and inner tubes are 2.54 cm and 1.27 cm, respectively. The gas enters the inner tube of the injector through two 0.63 cm ID tubes located at 10.79 cm below the top of the injector. These two tubes are situated at opposite sides of the injector. The gas then enters the column through a perforated plate at the top of the injector. There are seventeen 0.12 cm ID holes in the perforated plate. The liquid enters the column from the annulus of the injector.

Three types of solid particles are used in the experiments. They are 3 mm glass beads, 4 mm glass beads, and 2.5 mm nylon balls. The physical properties of these particles are given in Table 3.1 in Chapter 3. Water and air are used as the liquid and gas phases in the experiments. The superficial liquid and gas velocities in the experiments vary from 2.28 to 3.41 cm/sec and from 0 to 5 cm/sec,
Figure 4.1 Schematic Diagram of Gas-Liquid Injectors
respectively, based on the cross-sectional area of the outside column. A quick closing valve is employed to allow simultaneous shut-off of both the gas and liquid flows when measurement of the overall holdups of gas and liquid is desired.

RESULTS AND DISCUSSION

The results of the effects of various draft tube diameters and lengths, injectors, and solid particles under various gas and liquid velocities on the overall gas holdup for a draft tube gas-liquid-solid spouted bed under the circulated bed mode of operation (Fan et al., 1984b) are presented and discussed in the following:

Figure 4.2 shows the effect of draft tube diameter on the overall gas holdup and the ratio of the gas holdup to the superficial gas velocity with superficial liquid velocity as a parameter. The data points on the right hand side plot represent the maximum and minimum values of gas holdup/gas velocity ratio. Superficial liquid velocities of 2.28 cm and 3.41 cm are used in the experiments. Three draft tubes with 7.62, 10.16, and 12.7 cm ID are used, which yields ratios of the cross-sectional areas of the draft tube to the annulus as 0.38, 1.02, and 2.96, respectively. All of the three draft tubes are of identical length, i.e., 95.89 cm. Note that the overall gas holdup is primarily determined by the gas holdup in the draft tube region and that in the annular region. The gas holdup in the draft tube and the annular regions are strongly affected by the liquid
Figure 4.2 Effect of Draft Tube Diameter on the Overall Gas Holdup
circulation rate which varies with design and operational variables. The gas velocity in the draft tube region increases with an increase in the liquid circulation rate, resulting in a decrease in the gas holdup in the draft tube region. On the other hand, more bubbles are entrained into the annular region at higher liquid circulation rate, resulting in an increase in the gas holdup in the annular region. These two counteracting effects dictate the overall gas holdup in the draft tube gas-liquid-solid spouted bed. It is seen from Fig. 4.2 that among three draft tubes considered, the 7.62 cm ID draft tube has the highest overall gas holdup. Moreover, at a low gas velocity, the overall gas holdup for the 12.7 cm ID draft tube is higher than that for the 10.16 cm ID draft tube. However, the reverse is true at a high gas velocity. It should be noted that the overall gas holdup increases with an increase in gas or liquid velocity.

A draft tube spouted bed can be divided into four regions, i.e., below draft tube region, draft tube region, above draft tube region, and annular region (Fan et al., 1984b). The gas holdup in each region is normally different. The overall gas holdup for a draft tube spouted bed represents combined contribution of the local gas holdup from each of the four regions. The main effect of the bottom spacing (the vertical distance between the bottom edge of the draft tube and the conical section) is on the turbulence which occurs in the below draft tube region caused by the gas and liquid jets through the injector. Since the local gas holdup in the below draft tube region depends primarily on the turbulence in this region, it is affected by
the bottom spacing. Figure 4.3 shows the effect of the bottom spacing on the overall gas holdup, which reflects its effect on the local gas holdup in the below draft tube region. Four bottom spacings are considered, i.e., 0.44, 1.27, 2.54, and 5.08 cm. A draft tube of 10.16 cm ID is employed and the top spacing (the vertical distance between the top edge of the draft tube and the retaining grid) is fixed at 19.05 cm. At a liquid velocity of 2.28 cm/sec, the effect of the bottom spacing is negligible. Moreover, the overall gas holdup increases linearly with the gas velocity. The effect of the bottom spacing, however, is distinct at a liquid velocity of 3.41 cm/sec. The overall gas holdup for a bottom spacing of 1.27 cm (4 times the diameter of the 3 mm glass beads) is the highest. At a low gas velocity, a bottom spacing of 2.54 cm gives rise to the lowest overall gas holdup. However, at a high gas velocity, a bottom spacing of 0.44 cm gives rise to the lowest overall gas holdup. Note that in this figure a break point occurs at the gas velocity of about 2.8 cm/sec. The slope of the overall gas holdup versus the gas velocity below the break point is higher than that above the break point. This is due to the fact that the flow regime in the draft tube region changes from dispersed bubble flow regime to transition regime or slug flow regime.

The effect of the top spacing on the overall gas holdup at a fixed bottom spacing (2.54 cm) is shown in Fig. 4.4. Four top spacings are considered, i.e., 2.54, 5.08, 7.62, and 19.05 cm. The overall gas holdup decreases with an increase in the top spacing. A long top spacing refers to a short draft tube. Apparently, the bulk
Figure 4.3 Effect of the Bottom Spacing on the Overall Gas Holdup
Figure 4.4 Effect of the Top Spacing on the Overall Gas Holdup
density difference between the draft tube and the annular regions increases with an increase in the draft tube length. Consequently, the liquid circulation rate increases with a decrease in the top spacing, resulting in an increase in the overall gas holdup. The two counteracting effects described earlier need to be taken into account when comparing the overall gas holdup with a liquid flow to that without a liquid flow ($U_g = 0$). The bubble rise velocity in the draft tube region without a liquid flow is lower than that with a liquid flow. Hence, the gas holdup in the draft tube region without a liquid flow is higher than that with a liquid flow. However, the number of bubbles entrained into the annular region without a liquid flow is lower than that with a liquid flow. Hence, the gas holdup in the annular region without a liquid flow is lower than that with a liquid flow. Evidently, the overall gas holdup varies with the liquid velocity in complex functions of the top spacing and gas velocity as shown in Fig. 4.4. Note that in this figure the break point mentioned previously occurs at the same gas velocity (2.8 cm/sec). It is also noted that for a top spacing of 2.54 cm (the smallest), a large amount of liquid from the draft tube flows directly out of the retaining grid without recirculating to the annular region. This creates an empty section at the top of the annular region and results in an increase in the overall gas holdup measured. The bypassing of the liquid and the empty section are undesirable conditions for draft tube gas-liquid-solid spouted bed operation.
Figure 4.5 shows the effect of solid particle properties on the overall gas holdup. Particles considered include 2.5 mm nylon balls, 3 mm glass beads, and 4 mm glass beads. The total volume of the particles in the bed is fixed at 300 ml for all the particles used. It is seen in this figure that the overall gas holdup for 2.5 mm nylon balls is higher than that for 3 mm glass beads reflecting the particle density effects on the overall gas holdup. In addition, the overall gas holdup for 4 mm glass beads is higher than that for 3 mm glass beads. This may be due to the fact that bubble size for 4 mm glass beads is smaller than that for 3 mm glass beads. At a given gas velocity, the overall gas holdup increases with a decrease in bubble size. It is also seen in this figure that 2.5 mm nylon balls, which have the lowest particle terminal velocity among three particles used in the experiments, give rise to the highest overall gas holdup. Note that for 4 mm glass beads, at a liquid velocity of 2.28 cm/sec, the circulated bed mode (Fan et al., 1984b) could not be achieved for most of the gas velocities used in the experiments. Also, at a liquid velocity of 3.41 cm/sec, a packed bed is formed below the retaining grid when 2.5 mm nylon balls are used in the experiments. This flow mode cannot be classified as a circulated bed mode.

Figure 4.6 shows the effect of injector design on the overall gas holdup. The overall gas holdup for injector 1 is higher than that for injector 4 for most of the gas velocities employed in the experiments. This is due to an increase in bubble coalescence when the gas is introduced through the inner tube of a concentric injector, which
Figure 4.5  Effect of the Solid Particle Properties on the Overall Gas Holdup
Figure 4.6 Effect of the Injector Design on the Overall Gas Holdup
results in a decrease in the overall gas holdup. Therefore, when a concentric gas-liquid injector is used in a draft tube gas-liquid-solid spouted bed, it is recommended that gas is introduced into the column from the outer tube of the injector and liquid is introduced from the inner tube of the injector to prevent bubble coalescence and achieve a higher overall gas holdup. It can also be seen from this figure that the overall gas holdup increases with a decrease in the pore size of the distributor plate of the injector when a comparison of the overall gas holdup among injectors 1, 2, and 3 is made. The smaller the pore size is the smaller the bubble size will be, which results in an increase in the overall gas holdup. Injector 2 has the smallest pore size porous plate, thus, it gives rise to the highest overall gas holdup. However, it should be noted that the pressure drop across the distributor plate increases as the pore size of the distributor plate is decreased. Thus, in practical application, a compromise must be considered between the energy consumption and the overall gas holdup in order to optimize the bed performance.
CHAPTER 5

SOLID-LIQUID MASS TRANSFER IN A DRAFT TUBE
GAS-LIQUID-SOLID SPOUTED BED

The solid-liquid mass transfer plays an important role in determining the reactant conversion in the draft tube gas-liquid-solid spouted bed when it is used as a reactor for catalytic or non-catalytic multiphase reactions or a bioreactor for aerobic fermentation. Indeed, it is necessary to study the characteristics of the solid-liquid mass transfer in order to better quantify the overall reaction rate process in the system.

EXPERIMENTAL

A schematic diagram of the experimental apparatus is shown in Fig. 5.1. The apparatus is made of a 15.2 cm ID plexiglass tube 122 cm in height. A 10.2 cm ID plexiglass draft tube 96 cm in height is coaxially situated inside the column. The bottom of the column is a conical section with a cone angle of 90 degrees. The gas and liquid injector is located at the bottom of the conical section. Details of the injector design are given in Fig. 3.2 in Chapter 1. A movable
Figure 5.1 Schematic Diagram of the Experimental Apparatus.

1. Draft tube
2. Outside column
3. Injector
4. Particle withdrawing valve
5. Manometers
6. Upper retaining grid
7. Quick closing valves
8. Pressure gauge
9. Rotameters
10. Liquid pump
11. Liquid reservoir
12. Sampling valves
upper retaining grid, which serves to constrain the bed, is placed at the top of the column. In this study, the upper retaining grid is fixed at a location immediately underneath the fluid outlet. Thus, the total length of the bed, from the top of the injector to the upper retaining grid is 120.6 cm. For conventional gas-liquid-solid fluidized bed operation considered in this study, the draft tube spouted bed of present design is used, but without a draft tube. Furthermore, the conical section is replaced with a gas-liquid distributor whose design is given in Fan et al. (1982a).

Water and air are used as the liquid and gas phases in the experiments. The liquid and gas velocities vary from 0 to 2.11 cm/sec and from 0 to 5 cm/sec, respectively, based on the cross-sectional area of the outside column. A quick closing valve is employed to allow simultaneous shut-off of both the gas and liquid flows when measurement of the overall holdups of gas and liquid are desired. A water monometer at the bottom of the conical section measures the pressure at the bottom of the bed.

Cylindrical benzoic acid pellets are used as solid particles. These pellets are made by a 4 mm diameter punch. The height of the pellets is nearly equal to the pellet diameter. The wet density of benzoic acid pellets is 1.26 g/cc. After each experimental run had been completed, the benzoic acid pellets were withdrawn from the bed. These pellets were then reused after their diameter, height, weight, and density had been measured. The smallest size of benzoic acid pellets used in the experiments are 1.8 mm in diameter.
In-bed and outlet benzoic acid solution samples are taken from five sampling valves. A Bausch and Lomb Spectronic 2000 UV-visible spectrophotometer is used to obtain the concentration of benzoic acid solution.

RESULTS AND DISCUSSION

Benzoic acid solution, which acts as a surfactant, is known to have "surfactant effect" on the gas phase properties. For example, Levich (1962) and Keitel and Onken (1982) reported that the bubble size and bubble rise velocity decrease when a surfactant is present in the liquid phase. Zieminski (1967) found that bubble size decreases with an increase in surfactant concentration in the liquid phase. Lindland and Terjesan (1956), Gal-Or and Waslo (1968), and Raymond and Zieminski (1971) reported that bubble rise velocity decreases as the surfactant concentration increases.

Possible effects of solute (benzoic acid) concentration in the liquid phase on the solid-liquid mass transfer are evaluated in this study. The benzoic acid concentrations in the liquid encountered in the experiments range from 40 ppm to 450 ppm, which are very low compared to the saturated concentration, i.e., 2200 ppm. Figure 5.2 shows the effects of gas and liquid velocities, and solids (benzoic acid particles) loading on the overall gas holdup. It is seen that the overall gas holdup increases with an increase in gas or liquid velocity, which is consistent with that reported by
Figure 5.2 Effect of the Gas and Liquid Velocities and Solids (Benzoic Acid Particles) Loading on the Overall Gas Holdup in the Draft Tube Gas-Liquid-Solid Spouted Bed
Fan et al. (1984a, 1984b) who used pure water with inert particles of similar physical properties to benzoic acid particles. The overall gas holdup is seen to increase with an increase in solids loading, which is contrary to that reported by Fan et al. (1984a, 1984b). It is noted that this low range of benzoic acid concentration has induced a slight increase in the overall gas holdup over the pure water system. The overall gas holdup can be correlated with the gas and liquid superficial velocities based on the cross-sectional area of the outside column, solids loading, and particle diameter by the following equation:

$$
\varepsilon_g = 0.203 U_g \nu_g^{0.211} \nu_s^{0.118} d_p^{0.156}
$$

(33)

where $U_g$ and $U_g$ are in cm/sec, $\nu_s$ in ml, and $d_p$ in cm. Note that $d_p$ is particle equivalent diameter, which is equal to the diameter of a spherical particle that has the same volume as that of the cylindrical shape pellet used in the experiments.

The impact of slight variations of the gas holdup on solid-liquid mass transfer is evaluated by separate experiments in which a binary solid mixture is used to replace pure benzoic acid pellets as fluidized particles. This mixture is obtained by mixing a small amount (0.3% in volume) of benzoic acid pellets with a large amount (99.7% in volume) of inert PVC pellets, which have similar diameter and density to those of benzoic acid particles. The solid-liquid mass transfer coefficient obtained using this binary solid mixture is found to be within 5% of that obtained using pure benzoic acid pellets.
Thus, it is evident that hydrodynamic variation due to the presence of low concentrations of benzoic acid in the liquid do not establish significant effect on the solid-liquid mass transfer in the draft tube gas-liquid-solid spouted bed. Separate evaluations made on a conventional gas-liquid-solid fluidized bed by Arters and Fan (1984) have also concluded that the hydrodynamic properties of the solid particles in a conventional gas-liquid-solid fluidized bed were practically not affected by the presence of low concentrations of benzoic acid in the liquid phase.

The in-bed benzoic acid solution concentration profile for two operating conditions at a solids loading of 300 ml are shown in Fig. 5.3. The liquid velocities used are 0.52 and 2.11 cm/sec, and the gas velocities are 1.63 and 3.35 cm/sec. It is seen in this figure that the in-bed benzoic acid concentration is very uniform. Note that this is also the case when the gas is not present. Thus, the continuous flow stirred tank reactor (CSTR) model can be readily employed to evaluate the solid-liquid mass transfer coefficient for draft tube liquid-solid and gas-liquid-solid spouted beds, i.e., $k_2$ and $k_3$. The experimental solid-liquid mass transfer coefficient can be calculated by the following equation:

$$F(C_{\text{out}} - C_{\text{in}}) = k_i A (C_{\text{sat}} - C_{\text{out}}); \text{ } i = 2 \text{ and } 3 \quad (34)$$

where

- $F$ is liquid flow rate, cm/sec
- $C_{\text{in}}$ is inlet benzoic acid concentration, g/liter
- $C_{\text{out}}$ is outlet benzoic acid concentration, g/liter
- $C_{\text{sat}}$ is saturated benzoic acid concentration in water, g/liter
- $A$ is solid-liquid interfacial area, cm
Figure 5.3 Benzoic Acid Concentration Profile in the Gas-Liquid-solid Spouted Bed. L is the Downward Distance from the Upper Retaining Grid in the Spouted Bed.
Since the benzoic acid concentration in the inlet is equal to zero, Eqn. (34) becomes

\[ k_i = \frac{F_{\text{out}}}{A(C_{\text{sat}} - C_{\text{out}})} \quad ; \ i=2 \text{ and } 3 \] (35)

Although some small bubbles (0.1-2 mm) were observed to attach to the benzoic acid pellets, the portion of surface of benzoic acid pellets covered by these bubbles are negligibly small. Thus, the total surface area of benzoic acid pellets in the bed is used as the surface area in the calculation of the mass transfer coefficient. The slight change in particle shape due to rounding off of edges is taken into consideration in the calculation of particle surface area. A surface correction factor is defined as \((\text{true volume of particle}/\text{volume of perfect cylindrical particle})^{2/3}\). The true surface area is the surface area of perfect cylindrical particle times the correction factor.

Effects of liquid velocity on the solid-liquid mass transfer coefficient as a function of solids loading in the draft tube liquid-solid spouted bed are shown in Fig. 5.4. The gas velocity is set at zero in this case. It is seen in the figure that the mass transfer coefficient decreases with an increase in liquid velocity. In contrast, Rowe and Claxton (1965), Tournier et al. (1979), and Ballesteros et al. (1982) reported that the mass transfer coefficient in a conventional liquid-solid fluidized bed is independent of the liquid velocity for the gas Reynolds number ranging from 1.6 to 1320.
Figure 5.4 Effect of the Liquid Velocity on $k_2$ as a Function of Solids Loading in the Draft Tube Liquid-Solid Spouted Bed
A draft tube spouted bed can be divided into four regions, i.e., below draft tube region, draft tube region, annular region, and above draft tube region (Fan et al., 1984b). The solids holdup and mass transfer coefficient in each region are different. Moreover, they vary with liquid velocity. An overall mass transfer coefficient for a draft tube spouted bed represents combined contribution of the local mass transfer coefficient from each of the four regions. It was experimentally evident that an increase in the superficial liquid velocity increases the solids holdup, or mass transfer surface area, in the region above the draft tube and hence decreases the overall solids holdup in other three regions. The mass transfer coefficient in the region below the draft tube or the injector region is anticipated to be substantially higher than that in the region above the draft tube or in the other two regions, and to increase with an increase in the liquid velocity. Apparently, as the liquid velocity increases, the combined effects of the increasing mass transfer coefficient and decreasing surface area in the injector region yield a decrease in an overall mass transfer coefficient in the system. More experiments are needed to determine the local solids holdup and mass transfer coefficient in order to quantify their effects on the overall mass transfer coefficient. It is also seen in this figure that $k_2$ increases with an increase in solids loading.

Figure 5.5 shows the comparison of the solid-liquid mass transfer coefficient in a draft tube liquid-solid spouted bed with that in a conventional liquid-solid fluidized bed obtained from the correlation
Figure 5.5 Comparison of the Draft Tube Liquid-Solid Spouted Bed Data with Correlation of Ballesteros, et al. (1982) for the Conventional Liquid-Solid Fluidized Bed
of Ballesteros et al. (1982). It is seen that the solid-liquid mass transfer coefficient in the draft tube liquid-solid spouted bed is lower than that in the liquid-solid fluidized bed over the range of operating conditions used in this study. Note that the Sherwood number in the draft tube liquid-solid spouted bed can be described by a modified correlation of Ballesteros et al. (1982) as follows:

\[
\frac{Sh}{1.35 \phi^{0.323} Mv^{0.300} Sc^{0.400}} = 118.1 \epsilon_s^{0.283} Re_l^{-1.54}
\]  

(36)

where \( \epsilon_s \) is solids holdup and \( Re_l \) is liquid Reynolds number.

The effect of gas velocity on the solid-liquid mass transfer in the draft tube gas-liquid-solid spouted bed as functions of liquid velocity and solids loading is shown in Fig. 5.6. It is seen that the solid-liquid mass transfer coefficient increases with an increase in the gas velocity. This is consistent with that observed in the conventional gas-liquid-solid fluidized bed (Arters and Fan, 1984). The mass transfer coefficient in the draft tube gas-liquid-solid spouted bed increases with a decrease in the liquid velocity and with an increase in the solids loading. Similar observations were made in the draft tube liquid-solid spouted bed.

Figure 5.7 shows the effect of benzoic acid pellet diameter on the solid-liquid mass transfer coefficient. The gas and liquid velocity used are 3.36 cm/sec and 1.04 cm/sec, respectively. The mass transfer coefficient is seen to be independent of particle diameter for the particle diameter between 0.2 cm and 0.435 cm. This is
**Figure 5.6** Effect of the Gas Velocity on $k_3$ as Functions of Liquid Velocity and Solids Loading for a Draft Tube Gas-Liquid-Solid Spouted Bed

<table>
<thead>
<tr>
<th>$U_g$, cm/sec</th>
<th>0.52</th>
<th>1.04</th>
<th>2.11</th>
</tr>
</thead>
<tbody>
<tr>
<td>$V_s$, ml</td>
<td>150</td>
<td>O</td>
<td>△</td>
</tr>
<tr>
<td></td>
<td>300</td>
<td>⊙</td>
<td>△</td>
</tr>
</tbody>
</table>

$d_p = 2-4.5$ mm
Figure 5.7 Effect of Particle Diameter on $k_3$ for a Draft Tube Gas-Liquid-Solid Spouted Bed
consistent with that reported by Arters and Fan (1984) in their study of solid-liquid mass transfer behavior in a conventional gas-liquid-solid fluidized bed under comparable fluid velocity conditions.

Figure 5.8 shows that the solid-liquid mass transfer coefficient increases when gas is introduced into a draft tube liquid-solid spouted bed. This is due to the fact that the introduction of gas phase increases the turbulence and the liquid circulation rate in the draft tube spouted bed. Figure 5.8 also shows that the ratio of three phase/two phase Sherwood number can be correlated with gas and liquid Reynolds numbers in accordance with

\[
\frac{Sh_3}{Sh_2} = 1 + 1.088 \times 10^{-3} Re_L^{1.585} Re_g^{0.931}
\]  

(37)

where \(Sh_2\) and \(Sh_3\) are two phase and three phase Sherwood numbers, respectively. Figure 5.9 shows the comparison of experimental Sherwood number and that predicted by the above correlation equation. It is seen in this figure that the comparison is satisfactory.

The Sherwood numbers for the draft tube gas-liquid-solid spouted bed and the conventional gas-liquid-solid fluidized bed for various gas Reynolds numbers are shown in Fig. 5.10. The Sherwood number \(Sh_3'\) experimentally obtained for the gas-liquid-solid fluidized bed of 15.2 cm ID is found to be predictable by a correlation equation proposed by Arters and Fan (1984) as given by

\[
Sh_3' = 0.228(1+0.082Re_g^{0.623})_G a^{1.35} M_v^{0.323} Sc^{0.300} Sc^{0.400}
\]

(38)
Figure 5.8: Effects of Gas and Liquid Reynolds Numbers on Three Phase/Two Phase Sherwood Number Ratio for the Draft Tube System.
Figure 5.9 Comparison of the Experimental and Predicted Sherwood Number for Draft Tube Liquid-Solid and Gas-Liquid-Solid Spouted Beds
Figure 5.10 Comparison of Sherwood Number between the Draft Tube Gas-Liquid-Solid Spouted Bed and the Gas-Liquid-Solid Fluidized Bed
For comparison purposes, $Sh_{g}$ for the specific gas Reynolds numbers shown in the figure are presented based on the results calculated by Eqn. (38). Note that the correlation equation of Arters and Fan (1984) is established based on a 7.62 cm ID fluidized bed. For a low gas Reynolds number, it is seen that the Sherwood number in the draft tube gas-liquid-solid spouted bed is higher than that in the gas-liquid-solid fluidized bed at a low liquid Reynolds number. However, the former is lower than the latter at a high liquid Reynolds number. For a high gas Reynolds number, the Sherwood number in the draft tube gas-liquid-solid spouted bed is always higher than that in the gas-liquid-solid fluidized bed over the entire range of liquid Reynolds number used in the experiments.
CHAPTER 6

CONCLUDING REMARKS

In this study, experiments were conducted using various types of solid particles to investigate the hydrodynamic behavior of a gas-liquid-solid spouted bed with a draft tube. Three types of flow modes were observed, namely, the packed bed mode, the fluidized bed mode, and the circulated bed mode. Among these flow modes, the circulated bed mode is of most practical significance for application in the biological treatment of wastewater or other biological processes.

The pressure differences between the draft tube region and the annular region at the top and bottom levels of the draft tube constitute a driving force for liquid circulation from the draft tube region to the annular region. Based on the modified Fanning equation, a friction factor was defined to account for the friction loss in the bed. The friction factor varies solely with the Reynolds number defined based on the apparent liquid circulation rate. The bubble penetration depth in the annular region increases with an increase of the gas or liquid velocity. The bubble penetration depth, however,
decreases with an increase in solids loading.

The overall gas holdup in the bed increases with an increase in gas or liquid velocity but decreases with an increase in solids loading. The gas holdup in the draft tube gas-liquid-solid spouted bed is substantially higher than that in the conventional gas-liquid-solid fluidized bed at a given operating condition. The apparent liquid circulation rate increases as the gas or liquid velocity increases. At high gas flow conditions, there is an optimal solids loading under which the apparent liquid circulation rate is a maximum. The liquid circulation rate of the draft tube gas-liquid-solid spouted bed is higher than that of the gas-liquid airlift reactor at a high gas or liquid velocity. A model is proposed to describe the apparent liquid circulation rate satisfactorily.

The bubble size distributions in the draft tube and annular regions follow the log-normal distribution function. The average bubble size in the draft tube region is higher than that in the annular region for both the dispersed bubble regime and coalesced bubble regime in the draft tube region.

Experiments were also conducted to study the effect of draft tube size, top spacing, bottom spacing, injector design, and solid particle properties on the overall gas holdup in a draft tube gas-liquid-solid spouted bed. A draft tube of 7.62 cm ID gives rise to the highest overall gas holdup. At low gas velocities, the overall gas holdup for a 12.7 cm ID draft tube is higher than that for a 10.16 cm ID draft
tube. The reverse is true at high gas velocities. The effect of the bottom spacing is negligibly small at a low liquid velocity (2.28 cm/sec). At a high liquid velocity (3.41 cm/sec), the overall gas holdup for the four bottom spacings increases in the order 2.54 cm, 0.44 cm, 5.08 cm, and 1.27 cm for gas velocities below 4.2 cm/sec. The overall gas holdup increases with a decrease in top spacing.

The overall gas holdup for 2.5 mm nylon balls is the highest, followed by that for 4 mm glass beads. 3 mm glass beads give rise to the lowest overall gas holdup. A concentric injector with gas introduced from the outer tube of the injector yields better overall gas holdup performance than that with gas introduced from the inner tube of the injector. Moreover, a concentric injector attached with a porous gas distributor plate gives rise to a high overall gas holdup due to the small bubbles generated.

Finally, the solid-liquid mass transfer behavior in a draft tube gas-liquid-solid spouted bed of air, dilute benzoic acid solution, and benzoic acid pellets was studied. Continuous flow stirred tank reactor (CSTR) model was used to obtain the overall mass transfer coefficient. The mass transfer coefficient was found to increase with an increase in gas velocity or solids loading but decrease with an increase in liquid velocity. The mass transfer coefficient was found to be independent of particle diameter.
The mass transfer coefficients in the draft tube gas-liquid-solid spouted bed are higher than those in the draft tube liquid-solid spouted bed for a given liquid velocity. It was also found that the mass transfer coefficients in the draft tube liquid-solid spouted bed are lower than those in the liquid-solid fluidized bed for a given liquid velocity. For a low gas Reynolds number, the mass transfer coefficient in the draft tube gas-liquid-solid spouted bed is higher than that in the gas-liquid-solid fluidized bed at a low liquid Reynolds number. However, the former is lower than the latter at a high liquid Reynolds number. For a high gas Reynolds number, the mass transfer coefficient in the draft tube gas-liquid-solid spouted bed is higher than that in the gas-liquid-solid fluidized bed.

An empirical correlation developed by modifying the correlation of Ballesteros et al. (1982) for the liquid-solid fluidized bed was presented to portray the mass transfer coefficient in the draft tube liquid-solid and gas-liquid-solid spouted beds. This correlation was demonstrated to satisfactorily describe the solid-liquid mass transfer behavior in the draft tube liquid-solid and gas-liquid-solid spouted beds.
CHAPTER 7

RECOMMENDATIONS FOR FUTURE STUDY

Some recommendations for future study of the fundamentals of a draft tube gas-liquid-solid spouted bed are described in the following:

1. To study the jetting phenomenon in the below draft tube region and to determine its effect on the hydrodynamic and mass transfer behavior in the draft tube gas-liquid-solid spouted bed;

2. To develop methods to measure the true liquid circulation rate, i.e., the liquid circulation rate in the annular region;

3. To measure the local solid holdup and solid-liquid mass transfer coefficient in each of the four regions in the draft tube gas-liquid-solid spouted bed and to determine the contribution of the local mass transfer coefficient from each of the four regions to the overall solid-liquid mass transfer coefficient; and

4. To study the bed behavior involving small particles and to verify the applicability of correlations for large particle systems established in this study to small particle systems.
APPENDIX

COMPUTER PROGRAM FOR THE CALCULATION OF THE
APPARENT LIQUID CIRCULATION RATE IN A DRAFT
GAS-LIQUID-SOLID SPOUTED BED

* * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * *
* THIS PROGRAM CALCULATE THE APPARENT LIQUID CIRCULATION *
* RATE IN A DRAFT TUBE GAS-LIQUID-SOLID SPOUTED BED *
* * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * *

C
C INSIDE DIAMETER OF THE INJECTOR
DIN=1.899
C
C INSIDE DIAMETER OF THE DRAFT TUBE
D1=10.16
C
C OUTSIDE DIAMETER OF THE DRAFT TUBE
D10=11.43
C
C PI=3.1415926
C
C INSIDE DIAMETER OF THE OUTER COLUMN
D4=15.24
C
C BOTTOM SPACING
TLE=2.54/SQRT(2.)
C
C DRAFT TUBE LENGTH
RTL=37.75*2.54
C
C CROSS-SECTIONAL AREA OF THE OUTER COLUMN
A=(PI*D4**2)/4.0
C
C CROSS-SECTIONAL AREA OF THE INJECTOR
SIN=(PI*DIN**2)/4.0
C
C CROSS-SECTIONAL AREA OF THE DRAFT TUBE
\begin{align*}
S_1 &= \frac{(\pi D_1^2)}{4.0} \\
S_{10} &= \frac{(\pi D_{10}^2)}{4.0} \\
\text{CROSS-SECTIONAL AREA OF THE ANNULAR REGION} & \\
S_4 &= A - S_{10} \\
S_3 &= A \\
\text{TOTAL VOLUME OF THE BED} & \\
V &= 19360. \\
\text{VOLUME OF ABOVE DRAFT TUBE REGION} & \\
V_T &= 3475. \\
A_1 &= \frac{S_1}{S_4} \\
A_2 &= \frac{S_1}{S_3} \\
A_3 &= \frac{S_4}{S_3} \\
\text{INPUT PARAMETER : INLET LIQUID VELOCITY, GAS HOLDUP, LIQUID HOLDUP, AND SOLID HOLDUP IN THE DRAFT TUBE AND ANNULAR REGION, OVERALL LIQUID HOLDUP, LIQUID HOLDUP IN ABOVE DRAFT TUBE REGION} & \\
66 \quad \text{READ}(15,*,\text{END}=71) \ UIN,EPSSI,EPSLI,EPSCI, \ EPSSO, EPSLO, EPSGO, EPSLT, OLH \\
\text{WRITE}(6,10) \ UIN, EPSSI, EPSLI, EPSCI, EPSSO, EPSLO, EPSGO, EPSLT \quad \text{FORMAT}(/9F10.5/) \\
\text{WRITE}(6,45) \quad \text{FORMAT}(4X,'UIN',7X,'U1',7X,'U4'/) \\
C_1 &= UIN \cdot A / S_4 \\
\text{CALCULATE THE BULK DENSITY DIFFERENCE BETWEEN THE DRAFT TUBE REGION AND ANNULAR REGION} & \\
TPRE &= 980. \cdot RTL \cdot (2.52 \cdot (EPSSO-EPSSI)+(EPSLO-EPSLI)+1.2E-3 \cdot 5 \cdot (EPSGO-EPSCI)) \\
\text{SUCCESSIVE ITERATION METHOD} & \\
TU_4 &= 3 \cdot UIN \\
51 \quad \text{REY} &= \frac{(D_4-D_{10}) \cdot TU_4}{100.} \\
F &= 0.0791 / (\text{REY} \cdot 0.25) \\
FC &= 1.3066 \cdot F \cdot (S_4/SIN)^{**2} \\
B_1 &= 0.5 \cdot ((0.335+FC) \cdot (A_1^{**2})+(A_2-1.0)^{**2}-1.0) \\
B_2 &= -(0.335+FC) \cdot C_1 \cdot A_1 \\
B_3 &= 0.5 \cdot (0.335+FC) \cdot (C_1^{**2})+72.0 \cdot UIN \cdot TPRE}
\end{align*}
B4=B2**2-B1*B3*4.0
IF(B4.LT.0.) GO TO 31
U1=(-B2+SQRT(B4))/(2.0*B1)
U4=A1*U1-C1
IF(ABS(U4-TU4).LE.1.0E-3) GO TO 41
TU4=U4
GO TO 51

41 WRITE(6,42) UIN,U1,U4
42 FORMAT(3F10.5///)

C CALCULATE LIQUID CIRCULATION TIME IN EACH REGION
C
TD=RTL*EPSLI/U1
TA=RTL*EPSLO/U4
TT=VT*EPSLT/(U1*S1)
C
C TOTAL LIQUID CIRCULATION TIME
TTOTAL=TD+TA+TT
C
C CALCULATE APPARENT LIQUID CIRCULATION RATE
QL=V*OLH/TTOTAL
C
WRITE(6,75)
75 FORMAT(5X,'TD',7X,'TA',7X,'TT',9X,'TTOTAL'/)
WRITE(6,77) TD,TA,TT,TTOTAL
77 FORMAT(4F10.5///)
WRITE(6,76) QL
76 FORMAT(5X,'APPARENT LIQUID CIRCULATION RATE =',5X,F10.5)
GO TO 66

31 TU4=TU4+UIN
GO TO 51

71 STOP
END
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